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INTEGRATED DESIGN OF BIOLOGICAL PROCESSES AND THEIR CONTROL SYSTEM INCLUDING CLOSED LOOP PROPERTIES FOR DISTURBANCES REJECTION

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Abstract: This work presents an algorithmic approach to allow for the integrated design of processes and their control systems taking into account the controllability and the stability properties of the resulting system. The application of the proposed method has been carried out taking as an example model an alternative configuration to an Activated Sludge Process belonging to real wastewater treatment plant. In the integrated design, the process parameters are evaluated simultaneously with the parameters of the control system by solving a constrained non-linear optimisation problem. The considered cost function includes the investment and the operation costs and the constraints are selected to ensure that the values of some controllability and stability measures of the resulting system, are within specified ranges. These measures consist of the closed loop disturbances gains (CLDG) and the eigenvalues of some close-loop transfer matrices.

Keywords: Optimization problems, Processes, Process models , Controllability, Design.

1. INTRODUCTION

The traditional mode of designing processes has been the use of heuristic knowledge concentrated on determining the economically optimal process configuration among many possible alternatives. After the configuration is selected the process parameters and a stationary working point are evaluated by means of stationary models of the process in order to satisfy the operational requirements and to reduce investment costs. In this procedure, there is no consideration about the operability and controllability of the processes under design. The results have been plants very difficult to control and, consequently, in practice, there are a lot of self-controlled and very inflexible plants. The traditional approach to process control has been, given a designed process, to find the best selection and pairing of controlled and manipulated variables and also to find the controller parameters with the best closed-loop performance to work in a given operating point. The design and the control of processes were tasks performed sequentially; examination of controllability occurs only after the optimal process configuration and parameters are known.

This traditional methodology, however, ignores the ideas that changes in the process design might make the system more controllable (Luyben and Floudas 1994). Furthermore, better solutions can be found in the area of Integrated Design as shown in Gutiérrez (2000) and Luyben and Floudas (1994).

The main objective of the Integrated Design field is to obtain minimum cost designs while ensuring that some safety and performance constraints are satisfied for all possible plant parameters and disturbances. Research in Integrated Design aims to provide tools dealing with all aspects of the above objective in a well co-ordinated way to generate good designs efficiently.

In this paper, a systematic approach for the Integrated Design of Wastewater Treatment plants and their control system is presented. The parameters of the plant, the controller parameters and a stationary working point of the process are evaluated simultaneously while the investment and the operation costs are minimised. Constraints on some controllability measures (the closed loop disturbances gains (CLDG) and the eigenvalues of the state system transfer matrix) allow for a good dynamic resilience with respect to the disturbance rejection of the resulting closed loop system while the stability is guaranteed Skogestad, *et al*.(1997).

The proposed methodology consists of the combination of the design of the plant following a cost optimisation procedure, in addition with the desired closed loop dynamic as constraints. The independent variable set are the volumes of bioreactors, the settler cross sections, the gain and integral time of a PI controller and the working point.

The solution of such design problem involves the use of suitable algorithms for the non-linear constrained optimisation, a non linear mathematical model of the process and, also, linearised models that allow the closed loop transfer matrices and the controllability measures to be expressed in terms of the design parameters. The resulting optimal design (plant and controller) is tested in a non-linear simulation using real records of the disturbances taken from the real plant placed in Manresa, (Spain).

The paper begins describing the basis of the activated sludge process and the control objectives. An alternative configuration to the original plant and its control structure are also presented in the first section. The second section is devoted to the Integrated Design problem formulation and the methodology to be followed to get the optimal solution. In the last section, some results are commented in both, the time and the frequency domain, to end up with some conclusions.

2. DESCRIPTION OF THE ACTIVATED SLUDGE PROCESS AND CONTROL PROBLEM

As starting point, the alternative process represented in Fig. 2 to a the real plant located in Spain (Fig. 1), was selected. It consists of two aeration tanks working in series and two settlers. The basis of the process lies in maintaining a microbial population (biomass) into each bioreactor, transforming the biodegradable pollution (substrate) with dissolved oxygen supplied through aeration turbines. Water coming out each reactor, goes to the corresponding settler, where the activated sludge is separated from the clean water and recycled to both bioreactors. The control aim is to keep the substrate at the output, s_2, bellow a certain legal value despite the large variations of the flow rate and the substrate concentration of the incoming water (q_i and s_i). The whole set of variables are presented in Fig. 1 and 2. It can be noted that generically: “x” is used for the biomass concentrations (mg/l), “s” for the organic substrate concentrations (mg/l), “q” for flowrates (m3/h).

A first principle model of the system is obtained by considering the mass balances of oxygen, biomass and utilisation of organic substrate in the whole plant together with the equilibrium equations for the flows of water and sludge. Note that three layers of different and increasing biomass concentration are considered in the clarifiers. The control law corresponds to the well-known PI controller. Note that the controlled variable, in the alternative plant, is the substrate concentration in the second reactor,s_2, and the control signal is the flow rate of recycled sludge to the first reactor, q_{r1}, Gutiérrez (2000).

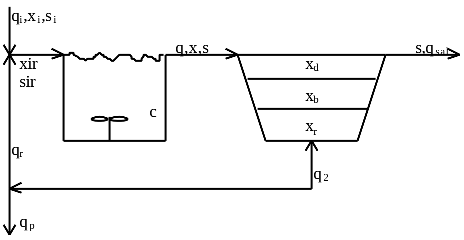

Fig 1. Original Plant

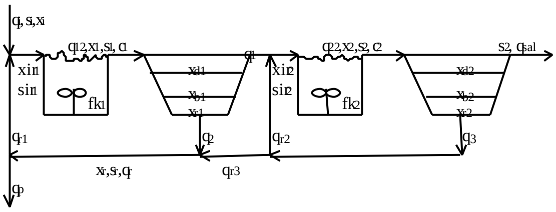

Fig 2. Alternative process for the real plant

3. THE OPTIMISATION PROBLEM FORMULATION

The Integrated Design problem for the alternative configuration consists of evaluating the physical dimensions of the process units, the controller parameters and a stationary working point simultaneously, while the investment and the operation costs together with the residuals associated to each equation of a non-linear first principle mathematical model of the process, are minimised subject to some physical and process constraints.

3.1 Cost function

The building costs are expressed in terms of the reactor volumes and the settler cross sections. The operation costs, given basically by the electrical energy consumed by the aeration turbines and by pumps, are expressed in terms of the magnitudes of the turbines aeration factors, fk_1, and fk_2 . The cost function is defined as:

$$J = \alpha r^2 + \beta V_1^2 + \chi V_2^2 + \delta A_1^2 + \mu A_2^2 + \varepsilon fk_1^2 , \quad + \phi fk_2^2 + \varphi r^2 \qquad (1)$$

The components of the vector of residuals, r, are given after normalisation by:

$$r(1) = \left(\frac{1}{40}\right)\begin{pmatrix} \mu y \dfrac{s_1 x_1}{k_s + s_1} - k_d \dfrac{x_1^2}{s_1} - k_c x_1 \\ + \dfrac{q_{12}}{v_1}(xir_1 - x_1) \end{pmatrix} \qquad (2)$$

$$r(2) = \left(\frac{1}{62}\right)\begin{pmatrix} -\mu \dfrac{s_1 x_1}{k_s + s_1} + f_{kd} k_d \dfrac{x_1^2}{s_1} \\ + f_{kd} k_c x_1 + \dfrac{q_{12}}{v_1}(sir_1 - s_1) \end{pmatrix} \qquad (3)$$

$$r(3) = \left(\frac{1}{31}\right)\begin{pmatrix} \mu y \dfrac{s_2 x_2}{k_s + s_2} - k_d \dfrac{x_2^2}{s_2} - k_c x_2 \\ + \dfrac{q_{22}}{v_2}(xir_2 - x_2) \end{pmatrix} \qquad (4)$$

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$$r(29) = \left(\frac{1}{4}\right)\begin{pmatrix} k_{la} fk_2 (c_s - c_2) - \mu y \dfrac{x_2 s_2}{k_s + s_2} \\ - \dfrac{q_{22}}{v_2} c_2 \end{pmatrix} \qquad (5)$$

3.2 Process constraints

1. *Residence times* in the aeration tanks and *mass loads* in the aeration tanks:

$$2.5 \le \frac{v_1}{q_{12}} \le 5 \ , \ 1.5 \le \frac{v_2}{q_{22}} \le 3 \qquad (6)$$

$$0.001 \le \frac{q_i s_i + q_{r1} s_r}{v_1 x_1} \le 0.06 \qquad (7)$$

$$0.001 \le \frac{q_1 s_1 + q_{r2} s_2}{v_2 x_2} \le 0.06 \qquad (8)$$

2. Limits in *Hydraulic capacity in the decanters* in the relatioship between the recycled and purge flow rates:

$$\frac{q_{12}}{A_1} \le 1.5, \ \frac{q_{22}}{A_2} \le 1.5, \ 0.5 \le \frac{q_2 + q_3}{q_i} \le 0.9 ,$$

$$0.03 \le \frac{q_p}{q_2 + q_3} \le 0.07 \qquad (9)$$

3. Limits in the *sludge ages* in the decanters

$$3 \le \frac{v_1 x_1 + A_1 l_{r1} x_{r1}}{q_p x_{r1} 24} \le 10 ,$$

$$3 \le \frac{v_2 x_2 + A_2 l_{r2} x_{r2}}{q_p x_{r2} 24} \le 10 \qquad (10)$$

3.3 Constraints on the process sensitivity gains for the main disturbances

The desired closed loop characteristics are good rejection of the two main disturbances (q_i and s_i) at the output variable, s_2, ensuring the stability of the resulting system. Mathematically the rejection of disturbances can be expressed as the ratio between the open loop and closed loop system disturbance gains (CLDG) and the stability as a set of constraints on the real parts of the eigenvalues (σ_2) of the closed-loop state matrix of the system, that should be all negative (Luyben and Floudas (1994), Mohideen, *et al.* (1995) and Skogestad, *et al.* (1997)).

By analysing time series of signals, taken from the real system, in the frequency domain it can be observed that the dominant frequencies are ω1=0.5 rad/h and ω2=0.1 rad/h. Consequently, the following constraints are proposed:

$$\rho_{s_2q_i}(0.5) < L_{sup},\ \rho_{s_2s_i}(0.5) < L_{sup}$$
$$\rho_{s_2q_i}(0.1) < L_{sup},\ \rho_{s_2s_i}(0.1) < L_{sup},\ \sigma_2 < 0 \quad (11)$$

where $\rho_{s_2q_i}(0.5), \rho_{s_2s_i}(0.5)$ are the ratios between the closed loop and open loop system gains of the output substrate (s_2) respect to the input flow (q_i) and respect to the input substrate (s_i), respectively, when they are evaluated at the frequency of 0.5 rad/h. Consequently, $\rho_{s_2q_i}(0.1), \rho_{s_2s_i}(0.1)$ are the same ratios of the system gains evaluated at 0.1 rad/h. L_{sup} are the upper bounds of those ratios and $\sigma 2$ denotes the real part of the eigenvalues of the closed loop state matrix of the system.

3.4 Control law

As mentioned above, the control law corresponds to the PI control and it is given by:

$$q_{r2}(t) = K_p((s_{ref} - s_2(t)) + \frac{1}{T_i}\int_0^t (s_{ref} - s_2(\tau))d\tau) \quad (12)$$

The problem is formulated in terms of the state space models of the process and the state space model of the controller. The nonlinear model and linearised model are both used. While the first one is used to obtain the operation point and the sensibility and the gains matrices, the linearised model is used to calculate the transfers function matrix and to express the controllability measures as functions of the design parameters. The solution was found numerically using the MATLAB software and its Optimisation Toolbox.

4. RESULTS OF THE OPTIMISATION PROBLEM

A set of results of the problem is shown in Table 1. The results of each design are the values for V_1, V_2 (volume of the aeration tanks), A_1, A_2 (cross areas of the clarifiers) and a set of steady state operating conditions defined by fk_1, fk_2 (aeration factors) and the substrate at the output of the plant, s_2. The last row shows the controller parameters which are the proportional gain, K_p, and the integral time, T_i.

In the second column, the results of one design, which was carried out without any restriction on the controllability measures, are presented. The third column and the following present results obtained with constraints on the controllability measures. The constraints are actually written in the first row of the table.

By comparison we can deduce that the building costs (given by V_1, V_2, A_1 and A_2) and the operation costs (given by fk_1,fk_2) are aproximately the same in all cases.

In the case of the design without constraints the reactors volumes are bigger than in the case of the design with the constraints $?_{s_2q_i}(0.1) < 0.9$. This shows that the plant dimension has an influence on its operability, flexibility and sensitivity respect to changes in the manipulated variables and the presence of disturbances.

Table 1. Results of the optimization problem (Closed loop designs)

	Without constraints	$\rho_{s_2q_i}(0.5) < 1$	$\rho_{s_2q_i}(0.1) < 0.9$
$\rho_{s_2q_i}(0.5)$	1.0029	1.	1.0340
$\rho_{s_2s_i}(0.5)$	1.0029	1.	1.0340
$\rho_{s_2q_i}(0.1)$	1.0150	1.	0.9
$\rho_{s_2s_i}(0.1)$	1.0150	1.	0.9
V_1	7310.	7042.	6504
V_2	7604.	7456.	4487.
A_1	1942.5	2068.	2067.
A_2	1073.1	1020.	2272.
f_{k1}	0.0498	0.0554	0.0607
f_{k2}	0.0051	0.0064	0.0185
s_2	22.1219	20.0143	20.9507
K_p	-0.3	-0.124	-7.52
Ti	1.2	18.	10.02
O. F.	1.8107	170.927	1.06189

However, the values of $\rho_{s_2q_i}(0.1), \rho_{s_2s_i}(0.1)$, $\rho_{s_2q_i}(0.5), \rho_{s_2s_i}(0.5)$ are reduced in some of cases in which constraints on them are considered. The working point also changes.

Some results can be seen in Fig. 4, 5 and 6. They are the time evolution of the substrate s_2 in the presence of disturbances in s_i obtained through the simulation of the non-linear model when the plant is operating in closed loop. In all the simulations real data records of disturbances, taken from the real process, were included. One of these records is actually, is presented in Fig. 3.

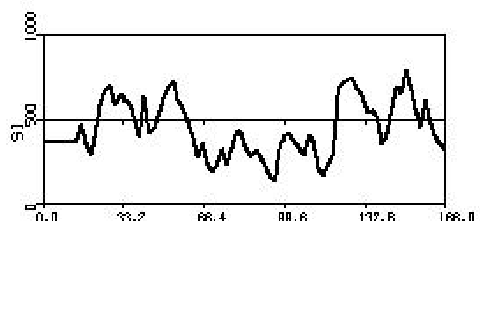

Fig. 3. Disturbance signal

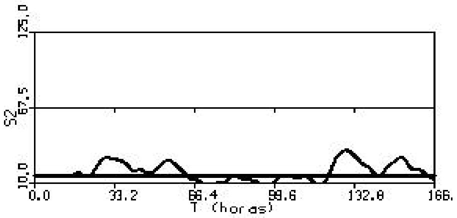

Fig. 4. Substrate at the output of the design with controllability constraints

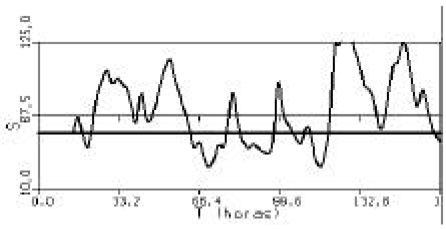

Fig. 5. Substrate at the output of the original plant

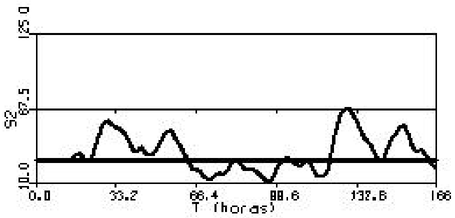

Fig. 6. Substrate at the output of the design without controllability constraints

Fig. 4 shows some simulation results for the substrate at the output, for the original plant when it was simulated in the presence of the disturbances of Fig. 3. The results of Fig. 5, which represents the substrate at the output of one designed alternative process, were obtained without imposing any constraints on the controllability measures in the Integrated Design optimisation problem. In contrast, with the results shown in Fig. 6 were obtained in the presence of such constraints. By comparing these two last figures we can say that the plant obtained by using integrated design techniques with constraints for the controllability measures, rejects the disturbance considerable better than the original plant and other designs obtained without constraints on them, as it was expected. The stationary value for s_2 is lower this implies a better operability of the process too.

5. ROBUSTNESS ANALYSIS

All the results above shown have been obtained by considering in the non-linear model the Monods aproximation for the biomass and the substrate kinetics in the reactors, i.e., $\frac{dx}{dt} = \mu x - k_d x$, $\frac{ds}{dt} = -\frac{\mu}{y} x$ being μ the microorganism growth which is expressed by $\mu = \mu_{max} \frac{s}{k_s + s}$. To verify the obtained models sensitivity respect to others dynamical models, the results obtained with the Monod model (the plant dimensions and the operatint point) were considered for three different non-linear simulations in which the Monod model , the Powell model , $\mu = \mu_{max} \frac{s}{((k_s + k_d) + s)}$, and the Haldane model, $\mu = \mu_{max} \frac{s}{\left(k_s + s + \frac{s^2}{k_s}\right)}$ were considered . The results are shown in Fig. 7, Fig. 8 and Fig. 9. All of them are very similar what shows the robustness of the obtained design (plant and controller)

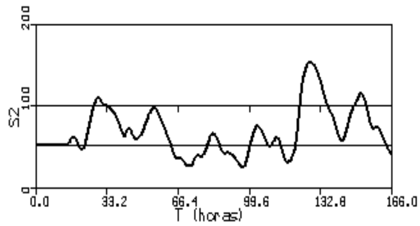

Fig 7. Modelo de Monod

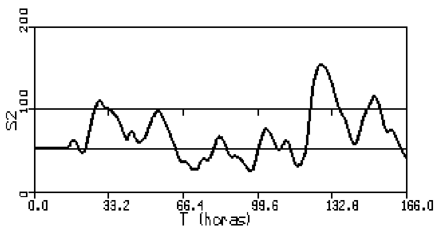

Fig 8. Powell model

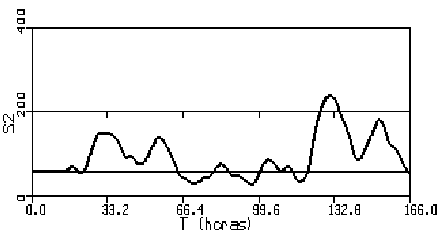

Fig 9. Haldane model

6. CONCLUSIONS

In this paper, the integrated design of activated sludge processes and of their control system has been formulated as a multiobjective constrained optimisation problem. It has also been solved for the design of an alternative configuration to a real plant placed in Spain together with the corresponding controller. The proposed approach allows us to include dynamic closed loop characteristics, through some controllability measures defined in the paper and physical and process operation constraints. The resulting designs have been proved to be better, in terms of the disturbances rejections at the dominant frequencies when some controllabillity measures are constrained while the costs (building and operation costs) are of about the same magnitude as in the designs carried out without those constraints. The robustness of the designs respect to others dynamical models has also been verified what is an important point from the practical point of view.

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